

**COMMERCIAL DEMONSTRATION OF THE NOXSO  
SO<sub>2</sub>/NO<sub>x</sub> REMOVAL FLUE GAS CLEANUP SYSTEM**

yes

DVORCAK
<i>[Signature]</i>
<i>[Signature]</i>

DOE/PC/90549--T/

RECEIVED

DEC 11 1996

OSTI

Contract No. DE-FC22-91PC90549

Quarterly Technical Progress Report No. 14

Submitted to

U.S. Department of Energy  
Pittsburgh Energy Technology Center

We have no objection from a patent  
standpoint to the publication or  
dissemination of this material.

*MPDVORCAK*  
Office of Intellectual  
Property Counsel  
DOE Field Office, Chicago

11/8/96  
Date

June 1, 1994 through August 31, 1994

Project Definition Phase

U.S. DOE Patent Clearance Is Not Required Prior to the Publication of this Document

MASTER

Prepared by

NOXSO Corporation  
2414 Lytle Road 3rd Floor  
Bethel Park, PA 15102

RECEIVED  
OCT 16 1996  
U.S.D.O.E.  
I.P.C.

DISTRIBUTION OF THIS DOCUMENT IS UNLIMITED

*Uw*

*743*

## **DISCLAIMER**

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

# **DISCLAIMER**

**Portions of this document may be illegible in electronic image products. Images are produced from the best available original document.**

---

## Table of Contents

<b>EXECUTIVE SUMMARY</b> . . . . .	<b>1</b>
<b>1 PROJECT DESCRIPTION</b> . . . . .	<b>3</b>
<b>2 PROCESS DESCRIPTION</b> . . . . .	<b>3</b>
<b>3 PROJECT STATUS</b> . . . . .	<b>6</b>
<b>3.1 Project Management</b> . . . . .	<b>7</b>
<b>3.2 NEPA Compliance</b> . . . . .	<b>7</b>
<b>3.3 Preliminary Engineering</b> . . . . .	<b>8</b>
<b>3.3.1 Separator Performance Test</b> . . . . .	<b>8</b>
<b>3.3.2 Construction Cost Estimating</b> . . . . .	<b>11</b>
<b>3.4 Nitrogen Oxide Studies</b> . . . . .	<b>17</b>
<b>3.5 Process Studies</b> . . . . .	<b>17</b>
<b>3.5.1 Laboratory Fluid-bed Adsorber Model</b> . . . . .	<b>17</b>
<b>3.5.2 Regenerator Computer Model Update &amp; Results</b> . . . . .	<b>23</b>
<b>3.5.3 Fluid Bed Flow Modeling</b> . . . . .	<b>28</b>
<b>3.5.4 SO<sub>2</sub> Production Processes</b> . . . . .	<b>36</b>
<b>3.6 Plant Characterization</b> . . . . .	<b>39</b>
<b>3.7 Site Survey/Geotechnical Investigation</b> . . . . .	<b>39</b>
<b>3.8 Permitting</b> . . . . .	<b>39</b>
<b>4 PLANS FOR NEXT QUARTER</b> . . . . .	<b>39</b>

## List of Figures

Figure 1-1	NOXSO Process Diagram . . . . .	4
Figure 3-1.	Separator Test Results Using POC Dust . . . . .	10
Figure 3-2.	Single-Stage Cold-Flow Adsorber Model . . . . .	19
Figure 3-3.	Grid Plate Design . . . . .	20
Figure 3-4	Three-Stage Adsorber Model . . . . .	21
Figure 3-5.	Comparison of the Simulated and Measured Sorbent Sulfur Content at the Regenerator Sorbent Inlet . . . . .	29
Figure 3-6.	Comparison of the Simulated and Measured Sorbent Temperature at the Regenerator Sorbent Inlet . . . . .	30
Figure 3-7.	Upper Section of Fluid-Bed Vessel . . . . .	31
Figure 3-8.	Centerline Velocity versus Height Above Top Fluid Bed . . . . .	32
Figure 3-9.	Centerline Velocity versus Height Above Top Fluid Bed . . . . .	33
Figure 3-10.	Dimensionless Centerline Velocity versus Dimensionless Height Above Fluid Bed . . . . .	35

### **List of Tables**

Table 3-1. Separator Test Results . . . . .	9
Table 3-2. Estimate Summary . . . . .	12
Table 3-3. Civil/Structural Estimate . . . . .	13
Table 3-4. Mechanical Estimate . . . . .	14,15
Table 3-5. Electrical Estimate . . . . .	16
Table 3-6. Estimate Scaling Factors . . . . .	18
Table 3-7. Velocities in Fluid-Bed Adsorber Test Apparatus . . . . .	22
Table 3-8. Grid Plugging Experimental Results . . . . .	22

## EXECUTIVE SUMMARY

The NOXSO process is a dry, post-combustion flue gas treatment technology which uses a regenerable sorbent to simultaneously adsorb sulfur dioxide ( $\text{SO}_2$ ) and nitrogen oxides ( $\text{NO}_x$ ) from the flue gas of a coal-fired utility-size boiler. In the process, the  $\text{SO}_2$  is converted to a sulfur by-product (elemental sulfur, sulfuric acid, or liquid  $\text{SO}_2$ ) and the  $\text{NO}_x$  is converted to nitrogen and oxygen.

The objective of the NOXSO Clean Coal Project is to design, construct, and operate a flue gas treatment system utilizing the NOXSO process at Alcoa Generating Corporation's (AGC) Warrick Power Plant. A project agreement between NOXSO and AGC was signed on August 30, 1994 enabling the project to be conducted at the Warrick Power Plant near Evansville, IN. The NOXSO plant is being designed to remove 98% of the  $\text{SO}_2$  and 75% of the  $\text{NO}_x$  from the flue gas from the 150-MW equivalent, unit 2 boiler. The by-product to be generated by the project is liquid  $\text{SO}_2$ . Sufficient construction cost and operating data will be obtained during the project to confirm the process economics and provide a basis to guarantee performance on a commercial scale.

The project is nearing completion of phase 1A, the project definition and preliminary design phase. The continuation application is being prepared to request DOE approval to proceed to phase 1B, Front End Engineering/Environmental Evaluation, of the project. Data obtained during pilot plant testing at Ohio Edison's Toronto Power Plant, which was completed on July 30, 1993 has been incorporated into the preliminary design of a commercial-size plant. This preliminary design is currently being updated for the conditions at Warrick Power Plant.

A draft EIV has been prepared by combining EIVs from the NOXSO project which was to be conducted at Ohio Edison's Niles Power Plant and the canceled CanSolv Project which was to be conducted at Warrick.

Preliminary engineering activities this quarter involved development of a construction cost estimating spreadsheet program. This spreadsheet has been used to estimate the cost of NOXSO systems at various sites. It will be updated during the plant construction to incorporate actual construction costs. The end product will be a quick, efficient, and accurate method to estimate the construction cost of a NOXSO plant. Additionally, laboratory experiments have been conducted on a module of the centrifugal separator to be used downstream of the NOXSO

adsorber to confirm vendor design data. Initial testing of the separator achieved lower removal efficiencies than claimed by the manufacturer. Although the results were lower than expected, the separator will be able to meet compliance for particulate emissions.

Process study activities include laboratory fluid-bed adsorber studies, regenerator computer model development and studies, fluid-flow modelling in fluid-bed vessels, and evaluations of SO<sub>2</sub> production processes. An initial study to determine proper grid spacing between adsorber beds has been completed. The laboratory-scale, fluid-bed adsorber will be used to conduct a study to improve the accuracy of the removal efficiency predictions and study the impact of adding a third adsorber stage. The regenerator computer model was developed this quarter and will be used to study design options for improving the regenerator performance. Fluid-flow modelling has been conducted to develop a procedure for determining the required vessel height above the top fluidized bed to prevent sorbent carryover. Evaluation of liquid SO<sub>2</sub> production processes is being conducted to determine the most economical method for producing liquid SO<sub>2</sub> for the project.

## 1 PROJECT DESCRIPTION

The objective of the NOXSO Demonstration Project (NDP), with cost-shared funding support from DOE, is to design, construct, and operate a commercial-scale flue gas cleanup system utilizing the NOXSO process. The NDP consists of the NOXSO plant and sulfur recovery unit, designed to remove  $\text{SO}_2$  and  $\text{NO}_x$  from flue gas and produce elemental sulfur by-product, and the liquid  $\text{SO}_2$  plant and air separation unit, designed to process the elemental sulfur into liquid  $\text{SO}_2$ . The NOXSO plant and sulfur recovery unit will be constructed at ALCOA Generating Corporation's (AGC) Warrick Power Plant near Evansville, Indiana, and will treat all of the flue gas from the 150-MW Unit 2 boiler. The elemental sulfur produced will be shipped to the Olin Charleston Plant in Charleston, Tennessee, for conversion into liquid  $\text{SO}_2$ .

The goals of the NDP include the reduction of the Warrick Power Plant Unit 2  $\text{SO}_2$  and  $\text{NO}_x$  emissions by 98% and 75%, respectively, and the waste minimization through the beneficial use of the sulfur by-product. In addition, construction costs and operating data from the project will be used to confirm the process economics and provide a basis to guarantee performance on a commercial scale. Ultimately, the successful demonstration of this process would assist utilities in attaining the emission limits specified by the 1990 Clean Air Act Amendments.

## 2 PROCESS DESCRIPTION

The NOXSO process is a dry, post-combustion flue gas treatment technology which will use a regenerable sorbent to simultaneously adsorb sulfur dioxide ( $\text{SO}_2$ ) and nitrogen oxides ( $\text{NO}_x$ ) from the flue gas from Unit 2 of AGC's Warrick Power Plant. In the process, the  $\text{SO}_2$  will be converted to liquid  $\text{SO}_2$  and the  $\text{NO}_x$  will be reduced to nitrogen and oxygen. The NOXSO plant is being designed to remove 98% of the  $\text{SO}_2$  and 75% of the  $\text{NO}_x$ . Details of the NOXSO process are described with the aid of Figure 1-1. Flue gas from the power plant is drawn through two flue gas booster fans which force the air through two-stage fluid-bed adsorbers and a baghouse before passing to the power plant stack. For simplicity, only one adsorber train is shown in Figure 1-1. Water is sprayed directly into the adsorber fluid beds as required to lower the temperature to 250-275°F by evaporative cooling. The fluid-bed adsorber contains active NOXSO sorbent, a 1.6 mm average diameter stabilized  $\gamma$ -alumina bead impregnated with sodium.  $\text{SO}_2$  and  $\text{NO}_x$  are adsorbed on the sorbent simultaneously by the following reactions:

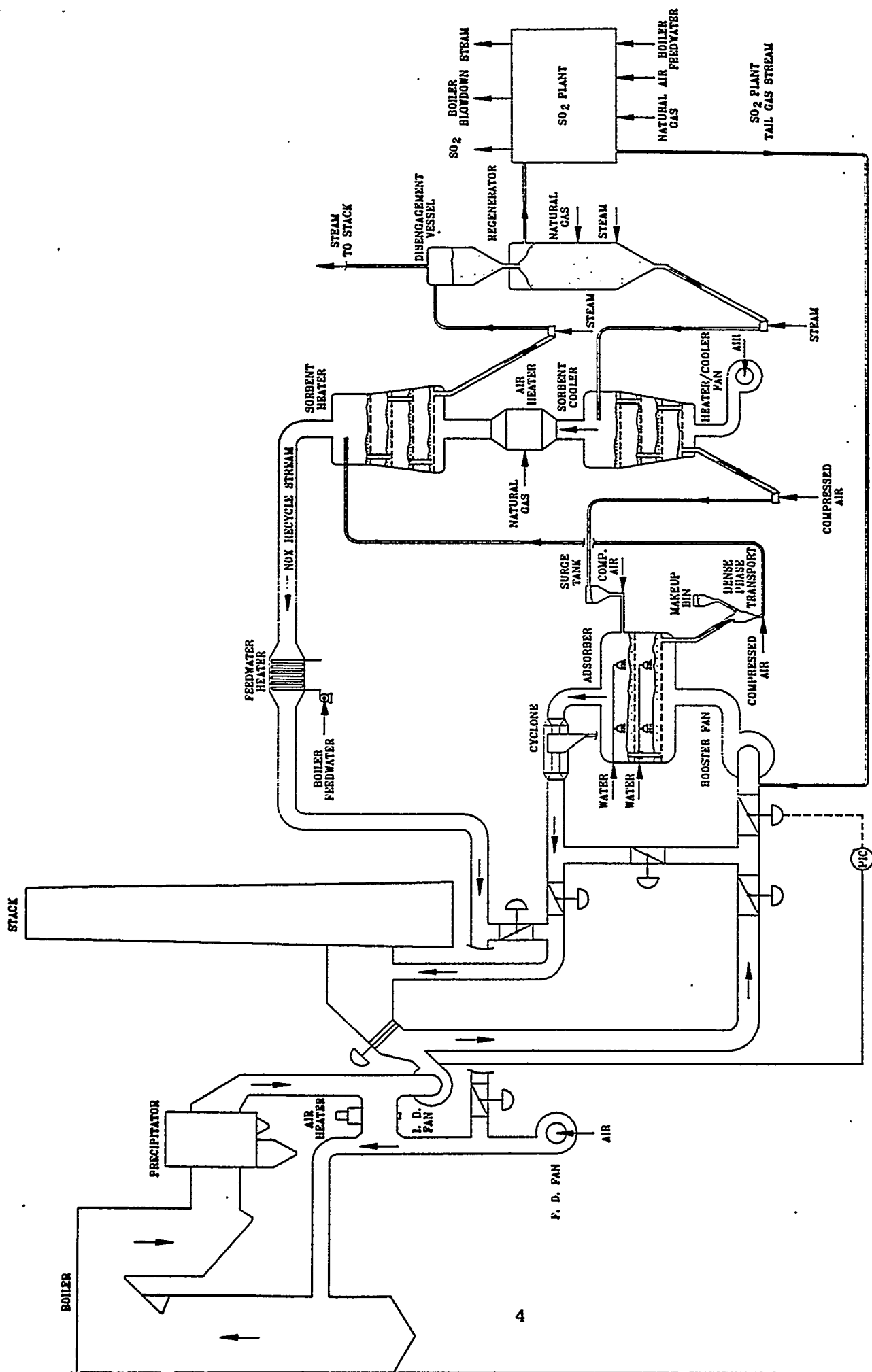
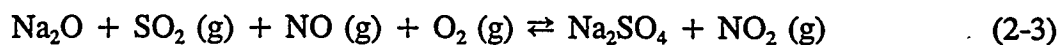
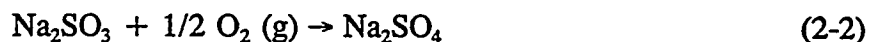
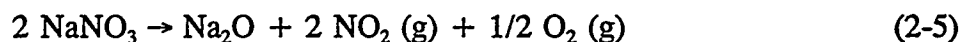


Figure 1-1. NOXSO Process Diagram



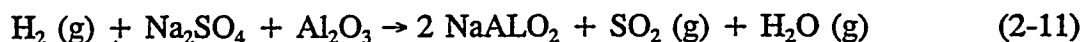
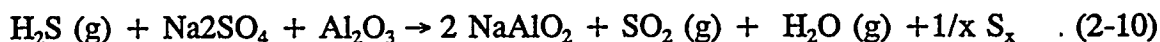
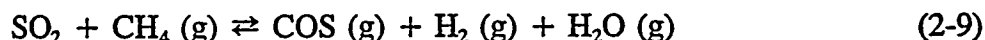
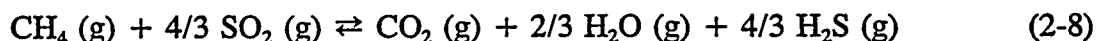
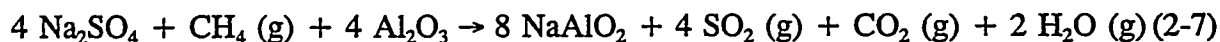
A baghouse separates sorbent which may be entrained in the flue gas and directs it to the fly ash sluicing system. Spent sorbent from the adsorbers flows into a dense-phase conveying system which lifts the sorbent to the top bed of the sorbent heater vessel. The sorbent flows through the four-stage fluidized bed sorbent heater in counterflow to the heating gas which heats the sorbent to the regeneration temperature of approximately 1150°F.

In heating the sorbent, the  $\text{NO}_x$  is driven off by the following reactions:



The evolved  $\text{NO}_x$  is then carried to the power plant boiler in the  $\text{NO}_x$  recycle stream. The  $\text{NO}_x$  recycle stream is cooled from approximately 360°F to 140°F in the feedwater heater. This heat-exchanger heats a slip stream of the power plant's feedwater, thereby reducing the amount of extraction steam taken from the low pressure turbine. The cooled  $\text{NO}_x$  recycle stream replaces a portion of the combustion air. The presence of  $\text{NO}_x$  in the combustion air suppresses the formation of  $\text{NO}_x$  in the boiler resulting in a net destruction of  $\text{NO}_x$ .

The heated sorbent is transported through an L-valve to the steam disengaging vessel. Transport steam is separated from the sorbent to reduce the volume of the off-gas stream. Sorbent gravity flows into the regenerator where it is contacted with natural gas. The sulfur on the sorbent combines with the methane and forms  $\text{SO}_2$  and  $\text{H}_2\text{S}$  by the following series of chemical reactions.





Additional regeneration occurs in the steam treater section of the regenerator when the sorbent is contacted with steam, converting the remaining sulfur on the sorbent to  $\text{H}_2\text{S}$ . The regenerator off-gas stream is directed to a sulfur recovery plant where the  $\text{H}_2\text{S}$  and  $\text{SO}_2$  are converted to elemental sulfur. Tail gas from the sulfur recovery plant will be oxidized and recycled back through the adsorbers to remove any residual sulfur compounds.

The elemental sulfur will be shipped to Olin Corporation where it will be oxidized to  $\text{SO}_2$  using a stream of oxygen. The  $\text{SO}_2$  vapor is then separated from other gaseous components, compressed, and liquified.

High temperature sorbent exiting the regenerator is conveyed with an L-valve to the four-stage fluidized-bed sorbent cooler. The sorbent flows counter to the ambient air which cools the sorbent. Regenerated sorbent exits the cooler at  $320^\circ\text{F}$ . The sorbent is then conveyed through an L-valve to the sorbent surge tank before being returned to the adsorber, completing the sorbent cycle.

Ambient air which is forced through the sorbent cooler by the heater-cooler fans exits the sorbent cooler at approximately  $950^\circ\text{F}$ . This preheated air then enters the air heater where it is heated to approximately  $1340^\circ\text{F}$  so it is capable of heating the sorbent exiting the sorbent heater to  $1150^\circ\text{F}$ .

### 3 PROJECT STATUS

The project definition and preliminary design phase, phase 1A, of the project is nearly complete. AGC's Warrick Power Plant has been selected as the host site for the project and a project agreement between NOXSO and AGC has been executed. The continuation application to the DOE to request continuation to phase 1B, Front End Engineering and Environmental Evaluation, is being prepared.

The NOXSO pilot-plant test program was completed on July 30, 1993. Performance at the pilot plant exceeded the initial expectations for pollutant removal efficiency, sorbent attrition, and electrical power and natural gas consumption. Data from the pilot plant has been incorporated into a fully integrated computer simulation which efficiently performs heat and material balances

for the combined NOXSO plant, power plant, and sulfur recovery plant. The computer program also calculates sizes and capacities for the major process equipment. This computer simulation is used to evaluate process alternatives to determine their impact on process economics.

A preliminary process flow diagram and associated heat and material balances have been prepared for a commercial-size plant. This flow diagram incorporates lessons learned from the pilot-plant test program as well as results of laboratory process studies, theoretical process studies, and the computer simulation. Preliminary piping and instrumentation diagrams have been prepared for a commercial-size plant based on the pilot plant experience and the preliminary process flow diagram.

A general arrangement has been prepared which incorporates plant design practices developed for the fluidized-catalytic-cracking (FCC) industry. Specifically, this design utilizes self supporting vessels, supported by skirts which extend from the vessel base to the foundation. In contrast, the POC design utilized a tower of structural steel on which the process vessels were mounted.

### **3.1 Project Management**

The Novation agreement to transfer the cooperative agreement from NOXSO to MK was executed on May 26, 1994. The effective date of the transfer is March 1, 1992.

The NOXSO/AGC project agreement was executed on August 30, 1994. This agreement defines the responsibilities of NOXSO and AGC and the business arrangement during the clean coal project and the following eight years of commercial operation.

All project management reports have been submitted to the DOE detailing project status, schedule, costs, and labor expended. Comparisons of planned and actual quantities were also submitted. Demolition of the pilot plant was completed on June 15, 1994.

### **3.2 NEPA Compliance**

With the selection of a host site for the NOXSO Demonstration Plant final work has begun to prepare the Environmental Information Volume. The Alcoa Generating Company's Warrick Power Plant was the host site for the canceled CanSolv project. A draft EIV for the CanSolv

project had been completed. Using the site specific information, from the CanSolv EIV and NOXSO Process specific information from the draft EIV for the NOXSO Demonstration Plant at Ohio Edison's Niles Power Plant, a preliminary EIV was developed.

The preliminary EIV was reviewed by DOE/PETC and their subcontractor to identify additional data needs required for NEPA compliance. A data needs list was developed and the information is being collected. As the data is obtained, the EIV is revised to reflect the additional information.

### **3.3 Preliminary Engineering**

#### ***3.3.1 Separator Performance Test***

The demonstration plant design employs centrifugal separators at the flue gas outlet from the adsorbers. The purpose of the separators is to remove attrited sorbent generated in the adsorbers from the exiting flue gas stream. The attrited sorbent is then directed to the boiler via the dense phase transport system, sorbent heater, and NO<sub>x</sub> recycle line. In the boiler, the attrited sorbent mixes with the ash and is removed with the bottom ash and fly ash. If the efficiency of the separators is high enough, they will remove a portion of the flyash from the flue gas which passes through the ESP, producing a net reduction in particulate emissions with the addition of the NOXSO plant.

The centrifugal separator selected is a straight through flow type. Using performance data provided by the separator manufacturer and attrited sorbent particle size distribution and loading data from the POC, a separation efficiency of greater than 60% is estimated. This would produce a significant reduction in particulate emissions since about a 30% removal efficiency is required to maintain the particulate emissions at the pre-NOXSO levels. To determine the actual efficiency, a test is being performed in the NOXSO lab. The test apparatus and set up were described in Quarterly Technical Progress Report No. 13 and some preliminary results are presented here.

The challenge dust being used for the tests was collected from the POC baghouse. The dust was fed into the gas stream at a rate producing a loading of about 0.07 grains/cubic foot. This particulate loading was determined from a typical PC fired boiler with a 99.1% ESP efficiency

and the latest NOXSO sorbent attrition data. Two test trials were conducted and the results are tabulated below.

**Table 3-1. Separator Test Results**

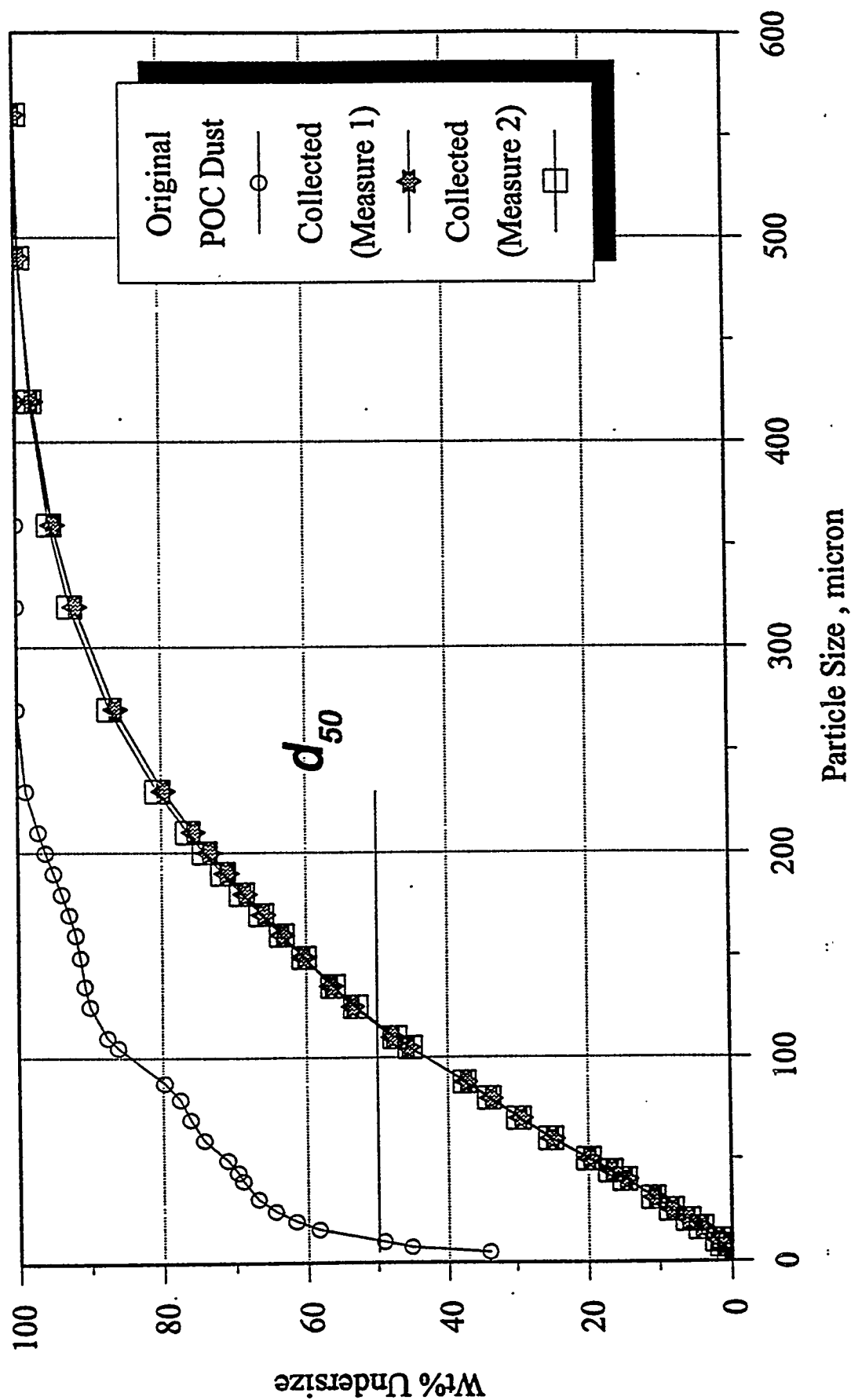
	Dust Input, g	Dust Collected, g	Removal Efficiency, %	Pressure Drop, "H <sub>2</sub> O
Trial 1	1362	492	36	2-3
Trial 2	1362	525	38	2-3

As seen, the separator removal efficiency was not as high as expected. In order to more completely explain the results, the particle size distributions of the challenge dust and the collected dust were measured. The results of this analysis are shown in Figure 3-1. There are two significant results reflected in this graph. First, the estimated removal efficiency was not achieved because the separator did not perform as expected on the smaller (sub-25 micron) sized particles. While manufacturers information indicates the separator would achieve about 90% removal for particles between 10 and 20 microns, the actual removal was closer to 30%. Because most of the mass of the challenge dust was less than 25 microns, the reduction in overall removal efficiency is a direct result of the lower than expected removal efficiency for sub-25 micron particles.

Also, the graph shows that there were larger particles in the collected, separated samples than were in the challenge dust. The two most likely causes are particle agglomeration and mass addition. Mass addition could be caused by weld slag or scale dislodging from the interior of the fan, duct, or separator during the test. To check for this possibility, a magnet was run through the original and collected samples to determine the metallic content. The results show that there was no metal in the original sample, while the collected samples averaged 2.5 wt% metal. This result reduces the reported removal efficiencies to 35% for Trial 1, and 37% for Trial 2. Meanwhile, Figure 3-1 suggests that approximately 10-15% of the collected particles were bigger than the original particles. This may be due to particle agglomeration, inconsistencies within the particle sizing instrument, and/or additional material which was not picked up by the magnet.

Figure 3-1. Separator Test Results

Using POC Dust



While the removal efficiency results are lower than expected, the separator performed as expected in terms of pressure drop. The 2-3" H<sub>2</sub>O pressure drop measured during the tests is a significant improvement over the typical 7-10" H<sub>2</sub>O of pressure drop associated with high efficiency cyclone separators. This represents a significant savings in both fan size and power consumption. Also, as indicated previously, it is believed that the separator will maintain compliance for particulate emission, in spite of the lower than expected removal efficiency. However, a particulate removal efficiency of greater than 35% is desirable and additional tests to increase the removal efficiency are planned. There are two modifications which can be made to the separator to increase its performance: adding a secondary collector, and adding an additional stage within the main collector.

In operation, the secondary system extracts a portion of the gas flow from the solid collection hopper of the primary collector. This fraction, about 10%, is then fed to a separate, secondary collector, particulate material is collected in a secondary hopper, and the gas is then fed back into the inlet of the primary collector for an additional removal cycle. The secondary flow, by drawing gas from the solid collection hopper of the primary separator, should assist the flow of sub-micron material from the separator into the collection hoppers. This set up is currently being tested in the lab. The second option, a two-stage separator, consists of two collectors in series within the primary collector. It has not yet been determined whether this separator will also be tested in the lab.

### *3.3.2 Construction Cost Estimating*

A computer spreadsheet for estimating the design and construction cost of a NOXSO plant has been developed. Table 3-2 shows an example of an estimate summary for the NOXSO plant at AGC's Warrick Power Plant. The estimate summary is separated into preliminary design, detail design, and construction. The costs for each phase of the project are further grouped into direct costs, indirect costs, and several miscellaneous cost categories.

Line item costs for equipment are organized by discipline as shown in Table 3-3 (Civil/Structural), Table 3-4 (Mechanical), and Table 3-5 (Electrical). The total line item costs

Table 3-2. ALCOA WARRICK PLANT  
BID SUMMARY

FROM ESTIMATE SUMMARY	CONSTRUCTION	PRELIMINARY DESIGN	DETAIL DESIGN	TOTAL
DIRECT LABOR MANHOURS	338752			338752
DIRECT COST				
CRAFT LABOR	\$6,329,103			\$6,329,103
PERMANENT MATERIAL	\$16,079,164			\$16,079,164
CONSTRUCTION EQUIPMENT				
COMPANY OWNED	\$196,088			\$196,088
OTHER	\$717,542			\$717,542
OPERATING EXPENSE	\$246,393			\$246,393
SMALL TOOLS AND CONSUMABLES	\$695,627			\$695,627
SUBCONTRACTS	\$17,644,852			\$17,644,852
SORBENT	\$1,738,682			\$1,738,682
OTHER COSTS "TEAM MEMBERS"	\$4,934,500	\$2,977,700	\$709,720	\$8,621,920
TOTAL - DIRECT COST	\$48,581,950	\$2,977,700	\$709,720	\$52,269,370
INDIRECT COST				
LABOR AND INSURANCE	\$3,304,308			\$3,304,308
SUPPLIES & EXPENSE	\$1,415,493			\$1,415,493
CONSTRUCTION EQUIPMENT	\$46,053			\$46,053
GENERAL PLANT	\$81,000			\$81,000
SUBCONTRACTS	\$82,128			\$82,128
OTHER COSTS (INCLUDING BOND & INSURANCE)	\$639,502	\$20,405	\$45,317	\$705,224
TOTAL - INDIRECT COSTS (BASED ON 21 MONTHS)	\$5,568,484	\$20,405	\$45,317	\$5,634,206
DESIGN ENGINEERING	\$703,095	\$2,320,295	\$2,022,900	\$5,046,290
ESCALATION (4 YR @ 2.9%/YR OF DIRECT & INDIRECT COST)	\$6,637,277	\$0	\$336,130	\$6,973,407
FINANCING				
	\$61,490,805	\$5,318,400	\$3,114,067	\$69,923,273
CONTINGENCY ( 0 % )	\$0			\$0
TOTAL PROJECT COST	\$61,490,805	\$5,318,400	\$3,114,067	\$69,923,273
MARGIN (G&A)	\$1,709,200	\$1,143,000	\$635,000	\$3,487,200
FEE	\$0			\$0
TOTAL ESTIMATE	\$63,200,005	\$6,461,400	\$3,749,067	\$73,410,473

## AT ALCOA W/ SO2 PLANT &amp; ESCALATION

Table 3-3. ALCOA WARRICK PLANT  
CIVIL/STRUCTURAL

DESCRIPTION	QUANTITY	UN	MAN HOUR	LABOR	EQUIP LABOR	PERM MATL	EQUIP RENT	EQUIP SUPPLY	SUPPLY & ST	SUBCONT	TOTAL
STRUCTURAL EXCAVATION	7939	CY	1708	\$57,128	\$12,957	\$0	\$16,530	\$21,119	\$11,591	\$0	\$119,325
TRENCH EXCAVATION	3000	CY	660	\$13,991	\$1,647	\$0	\$2,367	\$3,351	\$6,441	\$0	\$27,797
BACKFILL	3737	CY	2915	\$60,431	\$8,208	\$28,086	\$11,702	\$15,619	\$6,230	\$0	\$130,276
TRENCH BACKFILL	2700	CY	1372	\$30,247	\$594	\$1,501	\$1,534	\$1,328	\$6,237	\$0	\$41,441
CRUSHED ROCK SURFACING	1500	CY	426	\$9,437	\$714	\$24,877	\$1,275	\$1,949	\$779	\$0	\$39,030
DEMOLITION/REWORK EXISTING	1	LS	350	\$8,160	\$138	\$1,137	\$726	\$268	\$2,319	\$0	\$12,747
SITE GRADING	1	LS	219	\$4,614	\$793	\$0	\$1,240	\$1,272	\$476	\$0	\$8,396
ASPHALT PAVING	1012	SY	52	\$0	\$0	\$0	\$0	\$0	\$0	\$25,290	\$25,290
TOWER FILING	7956	LF	2793	\$0	\$0	\$0	\$0	\$0	\$0	\$198,912	\$198,912
PILE LOAD TEST	1	LS	0	\$0	\$0	\$0	\$0	\$0	\$0	\$35,000	\$35,000
DUCT SUPPORT FILING	7388	LF	2948	\$0	\$0	\$0	\$0	\$0	\$0	\$184,704	\$184,704
CONCRETE	5154	CY	24840	\$594,236	\$19,630	\$560,023	\$51,293	\$29,905	\$88,156	\$0	\$1,343,242
GROUTING	1	LS	182	\$4,233	\$0	\$1,137	\$0	\$0	\$873	\$0	\$6,242
CONC BLOCK WORK	5400	SF	1578	\$0	\$0	\$0	\$0	\$0	\$0	\$64,800	\$64,800
MAIN SUPPORT STEEL	166	TN	3198	\$88,011	\$3,430	\$216,573	\$21,075	\$7,078	\$9,073	\$0	\$345,242
MISC STEEL	28	TN	807	\$22,194	\$865	\$85,248	\$5,315	\$1,785	\$2,288	\$0	\$117,695
DUCT SUPPORT STEEL	193	TN	4700	\$127,704	\$6,608	\$251,197	\$31,414	\$11,745	\$13,166	\$0	\$441,835
GRATING	22505	SF	2251	\$61,278	\$4,366	\$192,084	\$12,288	\$8,169	\$6,324	\$0	\$234,510
COMPRESSOR BUILDING	1847	SF	1382	\$0	\$0	\$0	\$0	\$0	\$0	\$88,196	\$88,196
FIRE PUMP HOUSE	200	SF	163	\$0	\$0	\$0	\$0	\$0	\$0	\$10,400	\$10,400
ROOFING	4000	SF	360	\$0	\$0	\$0	\$0	\$0	\$0	\$26,000	\$26,000
TOUCHUP & LABELING	1	LS	672	\$14,966	\$878	\$2,200	\$449	\$794	\$1,543	\$0	\$20,830
MISC ARCH	1	LS	2179	\$51,667	\$516	\$25,000	\$1,483	\$656	\$5,327	\$0	\$84,649
ELEVATOR	1	EA	1661	\$0	\$0	\$0	\$0	\$0	\$0	\$108,000	\$108,000
STORM SEWER	300	LF	130	\$2,864	\$62	\$1,000	\$150	\$134	\$295	\$0	\$4,505
MIS/CE/SUMP	10	EA	208	\$4,605	\$99	\$5,000	\$241	\$214	\$475	\$0	\$10,634
ELECTRICAL MIS	3	EA	153	\$3,309	\$125	\$12,000	\$712	\$150	\$341	\$0	\$16,637
SIDING	34800	SF	1896	\$0	\$0	\$0	\$0	\$0	\$0	\$175,740	\$175,740
ASBESTOS REMOVAL	1	LS	487	\$0	\$0	\$0	\$0	\$0	\$0	\$20,000	\$20,000
TOTAL CIVIL/STRUCTURAL			60292	\$1,159,075	\$61,629	\$1,407,063	\$159,796	\$105,536	\$161,934	\$937,043	\$3,992,075

Table 3-4. ALCOA WARRICK PLANT  
MECHANICAL

DESCRIPTION	QUANTITY	UN	MAN HOUR	LABOR	EQUIP LABOR	PERM MATT	EQUIP RENT	EQUIP SUPPLY	SUPPLY & ST	SUBCONT	TOTAL
EQUIPMENT INSULATION	49910	SF	6040	\$0	\$0	\$0	\$0	\$0	\$0	\$494,604	\$494,604
PIPE INSULATION	23873	EF	5419	\$0	\$0	\$0	\$0	\$0	\$0	\$352,480	\$352,480
DUCT INSULATION	113229	SF	27000	\$0	\$0	\$0	\$0	\$0	\$0	\$1,122,102	\$1,122,102
AIR DRYERS	4	EA	149	\$3,897	\$131	\$140,616	\$557	\$140	\$401	\$0	\$145,743
AIR RECEIVERS	1	EA	44	\$1,169	\$39	\$5,933	\$167	\$42	\$120	\$0	\$7,469
AIR COMPRESSORS	4	EA	446	\$11,692	\$393	\$390,292	\$1,669	\$420	\$1,205	\$0	\$405,671
FG BFW PUMP	4	EA	149	\$3,897	\$127	\$13,885	\$549	\$138	\$401	\$0	\$18,998
DUCTWORK	688	TM	61150	\$1,739,089	\$62,026	\$1,385,403	\$249,981	\$59,503	\$179,288	\$0	\$3,675,290
ADSORBER BOOSTER FAN	2	EA	411	\$11,522	\$273	\$330,856	\$1,233	\$296	\$1,188	\$0	\$345,368
SORBENT STORAGE TANKS	2	EA	376	\$10,792	\$307	\$40,045	\$1,386	\$332	\$1,113	\$0	\$53,976
ADSORBER TANKS, INSTALL	2	EA	365	\$9,937	\$499	\$0	\$3,337	\$978	\$1,024	\$0	\$15,775
ADSORBER TANKS	2	EA	11512	\$0	\$0	\$0	\$0	\$0	\$0	\$2,273,776	\$2,273,776
SORBENT HEATER, INSTALL	1	EA	183	\$4,959	\$250	\$0	\$1,668	\$489	\$512	\$0	\$7,878
SORBENT HEATER	1	EA	5315	\$0	\$0	\$0	\$0	\$0	\$0	\$2,995,227	\$2,995,227
SORBENT COOLER, INSTALL	1	EA	183	\$4,959	\$250	\$0	\$1,668	\$489	\$512	\$0	\$7,878
SORBENT COOLER	1	EA	10545	\$0	\$0	\$0	\$0	\$0	\$0	\$1,110,161	\$1,110,161
SORBENT AIR HEATER	1	EA	62	\$1,742	\$59	\$125,722	\$463	\$149	\$179	\$0	\$128,315
NOx RECYCLE COOLER	1	EA	63	\$1,776	\$60	\$200,772	\$472	\$152	\$183	\$0	\$203,416
COOLING AIR FAN	3	EA	435	\$12,174	\$288	\$339,952	\$1,303	\$313	\$1,255	\$0	\$355,286
SORBENT SURGE TANKS	2	EA	356	\$10,205	\$290	\$21,458	\$1,310	\$314	\$1,053	\$0	\$34,631
SULFUR PLANT	1	LS	1970	\$51,227	\$1,689	\$4,579,567	\$7,059	\$1,825	\$5,280	\$0	\$4,646,649
CYCLONE	4	EA	976	\$26,528	\$1,334	\$469,065	\$8,907	\$2,611	\$2,735	\$0	\$511,180
DAMPERS & EXP JTS	55	EA	3446	\$98,812	\$2,807	\$351,966	\$12,681	\$3,049	\$10,186	\$0	\$479,501
DISENGAGEMENT VSL INSTALL	1	EA	183	\$4,968	\$256	\$0	\$1,682	\$494	\$512	\$0	\$7,912
DISENGAGEMENT VESSEL	1	EA	2403	\$0	\$0	\$0	\$0	\$0	\$0	\$431,821	\$431,821
DENSE PHASE VESSELS	6	EA	216	\$5,657	\$191	\$226,450	\$808	\$203	\$583	\$0	\$233,892
DENSE PHASE PIPING	670	LF	1424	\$37,372	\$1,205	\$0	\$5,226	\$1,304	\$3,852	\$0	\$48,960
SORBENT REGENERATOR, IN	1	EA	231	\$6,271	\$316	\$0	\$2,105	\$617	\$646	\$0	\$9,956
SORBENT REGENERATOR	1	EA	2403	\$0	\$0	\$0	\$0	\$0	\$0	\$431,821	\$431,821
REFRACTORY	1	LS	14822	\$0	\$0	\$0	\$0	\$0	\$0	\$1,703,931	\$1,703,931

### AT ALCOA W/ SO2 PLANT & ESCALATION

Table 3-4. ALCOA WARRICK PLANT  
MECHANICAL  
(CONT.)

[illegible]

## AT ALCOA W/ SO2 PLANT & ESCALATION

Table 3-5. ALCOA WARRICK PLANT  
ELECTRICAL

[illegible]

are the sum of labor, materials, construction equipment, construction supplies, and subcontract costs.

The basis for the cost estimate is the construction estimate prepared for the NOXSO plant which was to be installed at Niles. The NOXSO plant has been divided into subsystems and the appropriate cost scaling factor is applied to each item in the subsystem. The equation for calculating the "Scaling Factors" is given below. Table 3-6 lists the "base condition" for the base estimate and the "scale-up condition" for the example estimate. The calculated scaling factor and scaling exponents are also listed in Table 3-6.

$$\text{ScalingFactor} = \left[ \frac{\text{Scaleup Condition}}{\text{Base Condition}} \right]^{\text{ScalingExponent}}$$

### 3.4 Nitrogen Oxide Studies

No nitrogen oxide studies were conducted during this reporting period.

### 3.5 Process Studies

#### 3.5.1 Laboratory Fluid-bed Adsorber Model

A schematic of the first stage of the cold-flow, multi-stage, fluidized-bed adsorber (MSFBA) is shown in Figure 3-2. Details of the hole pattern for a grid plate with 2.1 % open area are shown in Figure 3-3.

The adsorber model shown in Figure 3-2 (cold) was constructed from plexiglass tubing and sheet. After construction, several shakedown tests were performed to verify the reliability of the equipment and identify any design deficiencies. Finally, tests were conducted to determine the required grid spacing to insure that no grid plugging would occur. The objective of this activity is to verify proper sorbent transport prior to constructing the three-stage adsorber model shown in Figure 3-4.

# AT ALCOA W/ SO2 PLANT & ESCALATION

Table 3-6. ALCOA WARRICK PLANT

## BASE CONDITION

PLANT CAPACITY	115 MW		SULFUR CONTENT	3 %
FLUE GAS FLOW	238526 SCFM		CIRCULATION RATE	262302 PPH
NO OF ADSORBERS	2 EA		NO OF HEATERS	1 EA
NO OF REGENS	1 EA		NO OF COOLERS	1 EA
BOOSTER FAN PWR	1826 HP		HTR/CLR FAN PWR	1160 HP
SORB. INVENTORY	627410 LBS		HTR/CLR AIR FLOW	268178 PPH

## SCALEUP CONDITION

PLANT CAPACITY	144 MW		SULFUR CONTENT	3.3 %
FLUE GAS FLOW	386000 SCFM		CIRCULATION RATE	386710 PPH
NO OF ADSORBERS	2 EA		NO OF HEATERS	1 EA
NO OF REGENS	1 EA		NO OF COOLERS	1 EA
BOOSTER FAN PWR	1950 HP		HTR/CLR FAN PWR	740 HP
SORB. INVENTORY	1159121 LBS		HTR/CLR AIR FLOW	399655 PPH

## SCALING FACTORS

CIVIL/STRUCURAL	1.14	EXP	0.40	GEN MECHANICAL	1.21	EXP	0.6
ELECTRICAL				ADSORPTION	1.33	EXP	0.6
INSTRUMENTS & CONTROL				REGENERATOR	1.26	EXP	0.6
				CIRCULATION	1.26	EXP	0.6
BOOSTER FANS	1.05	EXP	0.68	HTR/CLR FANS	0.74	EXP	0.68
SULFUR PLANT	1.27	EXP	0.62	NOx RECYCLE HX	1.29	EXP	0.65
HEATER	1.26	EXP	0.60	COOLER	1.26	EXP	0.6
DUCTWORK	1.42	EXP	0.40				

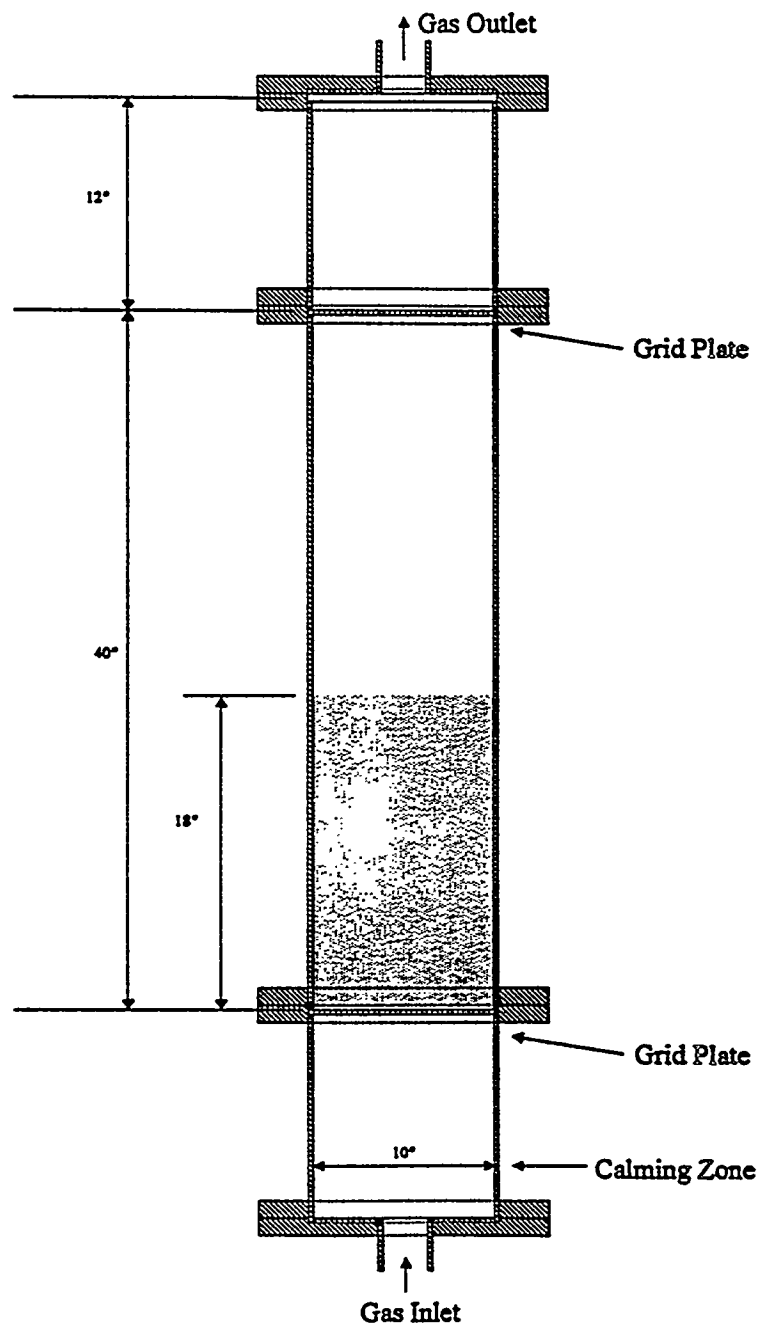
## VESSEL RATIOS

ADSORBER	1.00		HEATER	1.00
REGENERATOR	1.00		COOLER	1.00

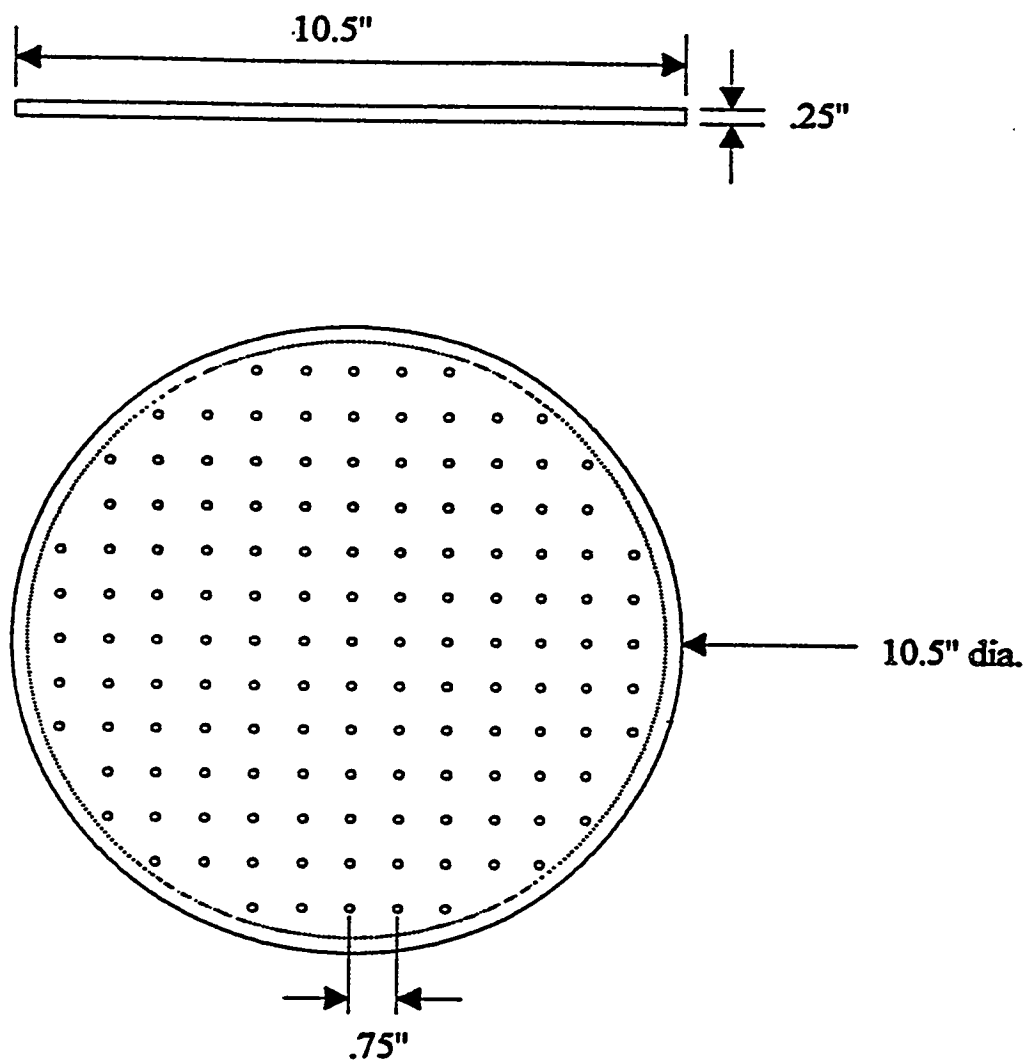
## ADJUSTMENT FACTORS

LABOR RATE FACTOR	0.97		
STRUCTURAL STEEL	0.2382		

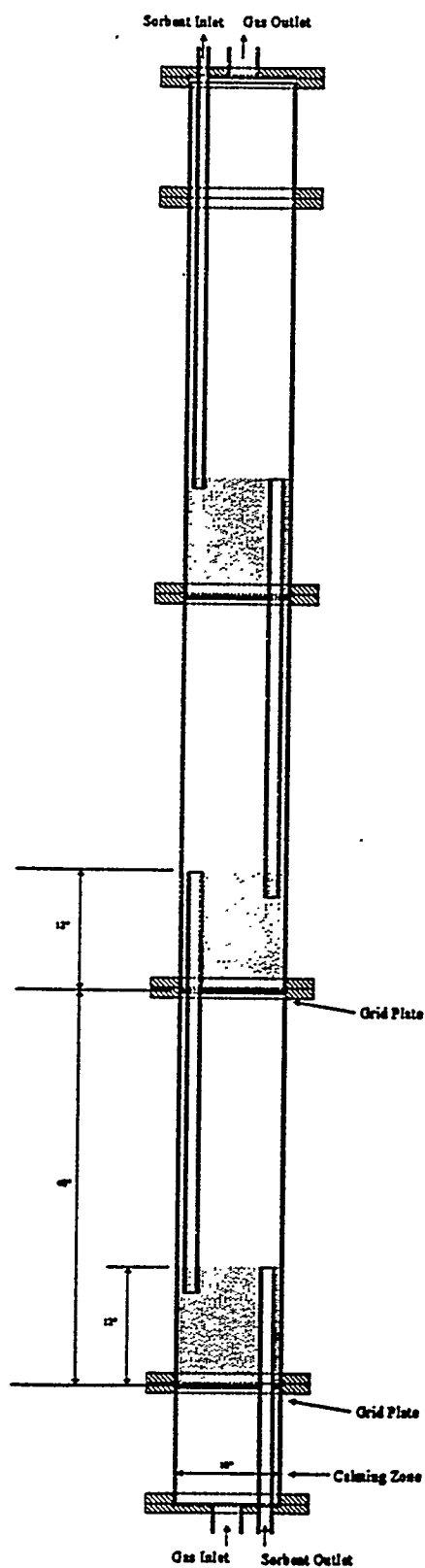
**Figure 3-2. Single-Stage Cold-Flow Adsorber Model**



**Figure 3-3. Grid Plate Design**



**Figure 3-4. Three-Stage Adsorber Model**



A pitot tube located in the inlet gas pipe was used to measure gas flow rate. The flow rate was adjusted to obtain a superficial gas velocity of 1.5 ft/s and a visual observation indicated the sorbent was at a state of minimum fluidization as expected. This provided a rough check on the flow measurement apparatus. Table 3-7 shows the cross sectional flow areas and gas velocities

**Table 3-7. Velocities in Fluid-Bed Adsorber Test Apparatus**

	Open Area (ft <sup>2</sup> )	Gas Velocity (ft/sec)	Gas Velocity (ft/sec)	Pitot Tube Measurement (H <sub>2</sub> O)	Gas Volume (ft <sup>3</sup> /min)
10" diameter (MSFBA)	.5454	1.5	90	-	91.63
		2.8	168	-	49.09
3" diameter (gas inlet)	.0491	16.67	1000	.05	91.63
		31.1	1866	.22	49.09

in the adsorber and inlet pipe at minimum fluidization (1.5 ft/s) and the design superficial velocity (2.8 ft/s). After it was confirmed that the required gas flow rate could be produced and measured, tests were conducted to determine the required grid to grid spacing as a function of sorbent bed depth to prevent grid plugging. As indicated in Table 3-8 tests were run with 18-

**Table 3-8. Grid Plugging Experimental Results**

	Bed Depth	Grid Spacing	Grid Plugging
Test 1	18"	40	yes
Test 2	12"	40	no
Test 3	18"	52	no

inch and 12-inch deep fluidized beds. The first two tests used a 40" tube length. In test 1 with the 18-inch deep bed, grid plugging occurred. The bed depth was reduced to 12 inches for the second test and no plugging occurred during a 30-minute test period. Finally, the top grid was raised 12" and test 3 was started. This test showed no plugging during a 30-minute test period.

### ***3.5.2 Regenerator Computer Model Update & Results***

In the last quarter, a regeneration reaction scheme was proposed to explain the observed phenomena. A computer program for the fixed-bed regeneration was also developed. To obtain the reaction rate constants, a least-squared method was used to match the computational results with the fixed-bed data. During this quarter, the computer program for the moving-bed regenerator was developed. The purpose of developing the moving-bed regenerator program is to improve the regenerator design. One example is to determine the inlet sorbent temperature. Since the sulfur regeneration is an endothermic reaction, the regeneration requires heat. With no heat source installed in the regenerator, the reaction consumes the sorbent's sensible heat causing the temperature to drop as the sorbent moves toward the outlet. To ensure the regeneration occurs throughout the regenerator, one has to set the sorbent-inlet temperature high. But from the energy and material point of view, we intend to lower the temperature. With this program, we can quickly predict the required sorbent inlet temperature, as a function of operating parameters.

However, the assumptions made in the mathematic model limit the use of the computer program. Furthermore, the lack of the laboratory data forces us to several assumptions to the calculation. Therefore, one should be aware of these limitations when applying the simulation results to the regenerator design. The program was developed to estimate the temperature, pressure and concentrations along the regenerator. The program requires the following information to start its calculation.

1. sorbent flow rate
2. regenerant gas flow rate
3. spent sorbent sulfur content
4. regenerated sorbent sulfur content
5. inlet methane temperature
6. sorbent temperature at the regenerant inlet
7. sorbent properties, such as surface area, mean particle size, particle sphericity, particle density, particle loosely-packed bulk density.
8. reactor diameter
9. ambient temperature

Depending upon the choice of the gas-solid flow pattern, the program reports the results for either co-current or counter-current regeneration. At the end of calculation, the program gives the following information along the regenerator.

1. gas concentrations
2. sorbent sulfur content
3. gas temperature
4. sorbent temperature
5. pressure
6. gas flow rate
7. sorbent residence time
8. sorbent inventory

### Model and Assumptions

#### *Reaction Chemistry*

Based on the regeneration test studies, a reaction scheme consisting of seventeen reactions was proposed in the last quarterly report to explain the regeneration of the spent NOXSO sorbent. The same reaction scheme was used in the moving-bed regenerator model.

#### *Governing Equations*

Both W.R. Grace Research Center and NOXSO Laboratory proved that the regeneration is not diffusion limited. A one-dimensional flow model with no diffusion resistance was used to simulate the moving-bed regenerator. The assumptions are listed in the following.

1. no external or internal diffusion resistance
2. constant reaction heat
3. constant gas viscosity
4. constant gas and solid heat capacities
5. uniform particle temperature
6. heat generation in the solid phase
7. heat loss occurs from the gas phase

The assumptions made in number 1 through 5, simplify the model which results in a faster computer program. The modified Ergun equation is used to model the pressure drop in the moving-bed reactor. The governing equations are listed in the following.

Gas phase

$$\frac{d}{dW}(F_g C_i) = SA \, r_i$$

Solid phase

$$\frac{d}{dW}(F_s C_{sj}) = SA \, r_{sj}$$

Pressure

$$\frac{dP}{dW} = \frac{(1-\epsilon)}{\epsilon \phi_s d_p g_c A^2 \rho_B} \left( \frac{150(1-\epsilon)}{\epsilon \phi_s d_p} \mu \Delta F_g + \frac{1.75}{A} \rho_g \Delta F_g^2 \right)$$

Gas temperature

$$\frac{d}{dW}(F_g \rho_g C_{pg} T_g) = -6 \frac{h_g}{d_p \rho_s} (T_g - T_s) - \frac{4}{D} \frac{U}{\rho_B} (T_g - T_a)$$

Solid temperature

$$\frac{d}{dW}(F_s C_{ps} T_s) = -6 \frac{H_g}{d_p \rho_s} (T_s - T_g) + \sum (-\Delta H_i) SA \, r_i$$

where

- A = reactor cross-sectional area, cm<sup>2</sup>
- C<sub>i</sub> = i-th gas concentration, gmole/cm<sup>3</sup> reactor
- C<sub>pg</sub> = averaged gas heat capacity, cal/gmole.°C
- C<sub>ps</sub> = solid specific heat, cal/g.°C

$C_{sj}$	= j-th solid concentration, gmole/g sorbent
$D$	= reactor diameter, cm
$F_g$	= gas flow rate, cm <sup>3</sup> /sec
$g_c$	= gravitational acceleration, 980 cm/sec <sup>2</sup>
$h_g$	= solid to gas heat transfer coefficient, cal/cm <sup>2</sup> .sec.°C
$P$	= pressure, atm
$r_i$	= i-th gas reaction rate, gmole/sec.m <sup>2</sup> sorbent
$r_{sj}$	= j-th solid reaction rate, gmole/sec.m <sup>2</sup> sorbent
$SA$	= sorbent surface area, m <sup>2</sup> /g
$T_a$	= ambient temperature, °K
$T_g$	= gas temperature, °K
$T_s$	= solid temperature, °K
$U$	= heat transfer coefficient, cal/cm <sup>2</sup> .sec.°C
$W$	= sorbent weight, g
$\Delta F_g$	= gas flow rate relative to the solid flow, cm <sup>3</sup> /sec, = $F_g \pm F_s/\rho_s$
$\Delta H_i$	= Heat of i-th reaction
$\varepsilon$	= reactor voidage
$\mu$	= gas viscosity, g/cm.sec
$\rho_s$	= particle density, g/cm <sup>3</sup>
$\rho_B$	= particle loosely packed bulk density, g/cm <sup>3</sup>
$\rho_g$	= gas molar density, gmole/cm <sup>3</sup>
$\varphi_s$	= particle sphericity

The corresponding initial conditions are these known values of gases and solids at the gas inlet. A 5-th order Runge-Kutta method with step-size control routine was used to integrate the coupled differential equations.

#### Values of Constant Properties Used in the Simulation Program

##### *Regeneration Heat*

$\Delta H = 942$  cal/g sulfur (Determined by curve-fitting the POC regenerator temperature profiles)

### *Heat Transfer Coefficient between Gas and Sorbent Particles*

$$h_g = 4.e^{-4} \text{ cal/sec.cm}^2.\text{°C} \text{ (Based on PETC batch fluid-bed heater and POC 3-stage fluid-bed cooler modeling results)}$$

### *Heat Loss Coefficient*

$$U = 9.033e^{-6} \text{ cal/sec.cm}^2.\text{°C} \text{ (Based on Babcock \& Wilcox Kaowool Ceramic Fiber Products, Blanket B 1800)}$$

### *Solid Specific Heat*

$$C_{ps} = (22.08 + 0.008971 \cdot T_s - 533500/(T_s^2 - 625))/102, \text{ cal/g.°K} \text{ (mean solid specific heat of solid temperature at gas inlet } \pm 25\text{°C)}$$

### *Gas Heat Capacity*

$$C_{pg} = 6.557 + 0.001477 \cdot T_g - 0.02148 \cdot 10^{-6} \cdot T_g^2 + 625/3, \text{ cal/gmole.°K} \text{ (mean air heat capacity at gas inlet temperature } \pm 25\text{°C)}$$

### *Gas viscosity*

$$\mu = 3.72e^{-4} \text{ g/cm/sec}$$

### *Ambient Temperature*

$$T_a = 20\text{°C}$$

### Curve-fit POC Moving-Bed Regenerator Data

The POC regenerator is a non-adiabatic reactor with gas and solid flowing in the counter-current direction. A nitrogen purge was used to prevent instrument sensing tubes from being clogged by the sorbent and sulfur deposits. Steam flows were introduced into the reactor through the top-J valve and steam treater. The off-gas concentrations obtained from the POC regenerator have many unmeasured quantities. Therefore, we only used the sorbent sulfur-content data to

determine the temperature dependence of the regeneration rate. According to the Arrhenius' law, the reaction rate has exponential temperature dependence. The curve-fit calculation becomes very unstable creating numerical problems with the results. To obtain a workable result, we added another assumption.

*All reaction rates have the same temperature dependence.*

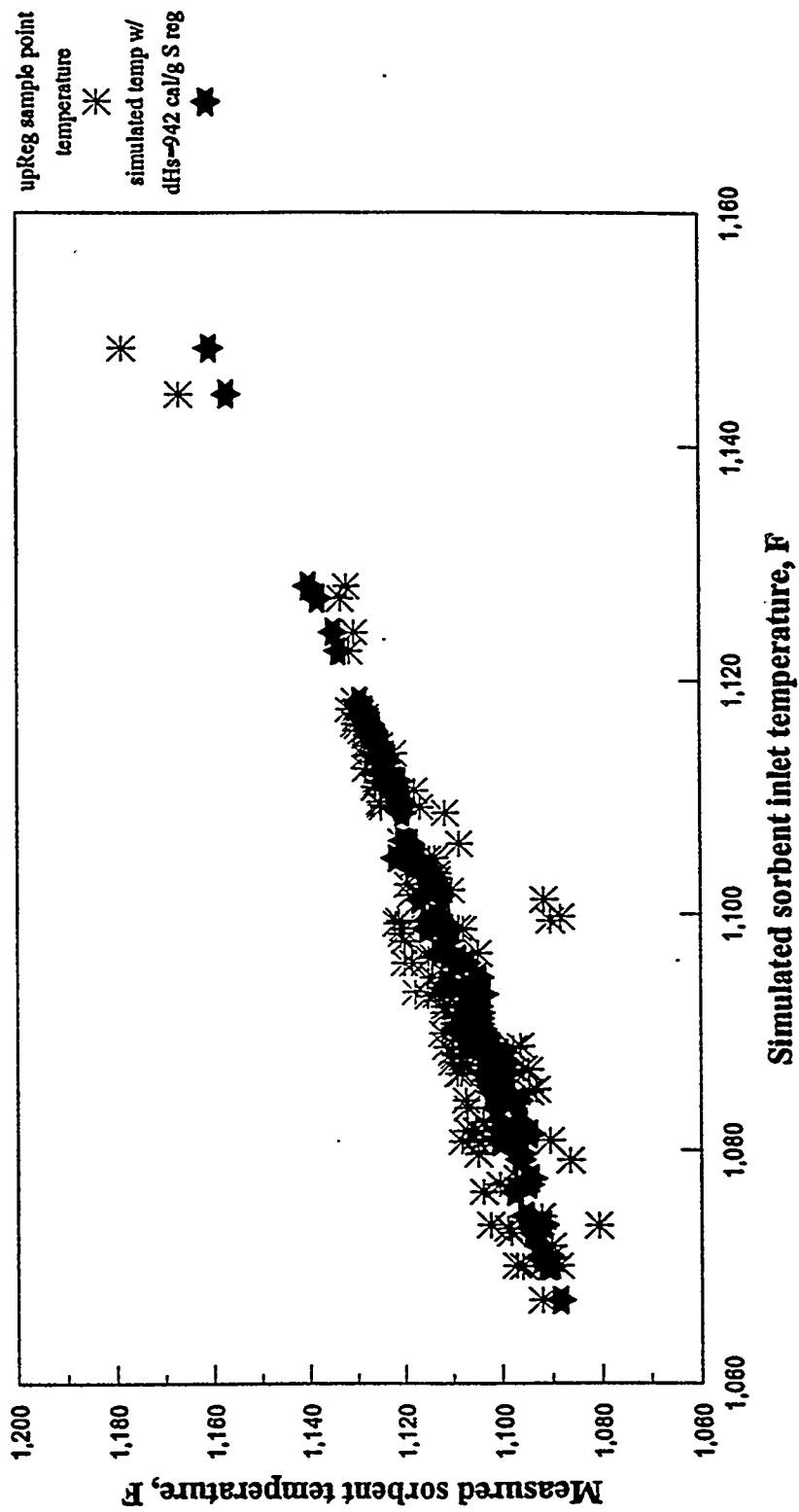
Instead of searching for 17 activation energies for the entire reaction scheme, we are looking for two parameters. One is the common activation energy, and the other is the mean regeneration heat. The two best-fit parameters were used to simulate the sorbent sulfur content to verify the result. For the counter-current flow, we used the sorbent-outlet sulfur content and temperature to calculate the same parameters at the sorbent inlet. The results are shown in Figure 3-5 and Figure 3-6. The former is for the comparison of sorbent sulfur content, and the latter is for the sorbent temperature. Both figures show the fits being satisfactory.

### **3.5.3 Fluid Bed Flow Modeling**

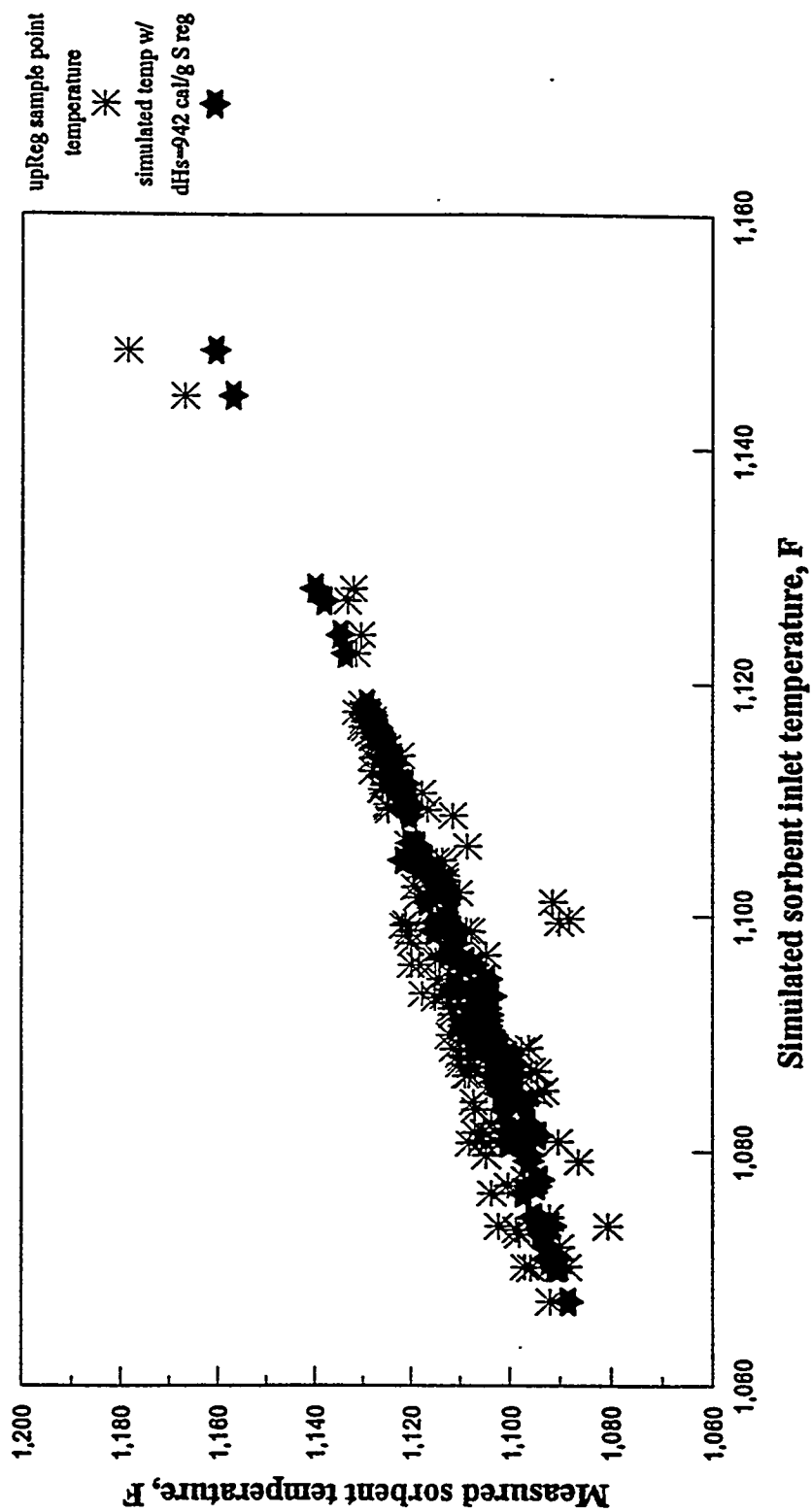
Flow modelling of gas exiting a fluidized bed vessel was performed. The vessel arrangement and nomenclature used in subsequent discussions are given in Figure 3-7. The objective of the flow modelling is to determine the effect of the vessel height above the grid (H) and the vessel diameter (D) on the velocity field above the fluidized bed. Specifically, the vessel design must assure that the "gas acceleration zone" does not encroach into the "transport disengaging zone". The transport disengaging zone is the region where sorbent which "splashes" above the fluid bed due to bursting gas bubbles can fall back into the bed. High gas velocities in this region would cause sorbent to be carried out of the vessel, and must be avoided.

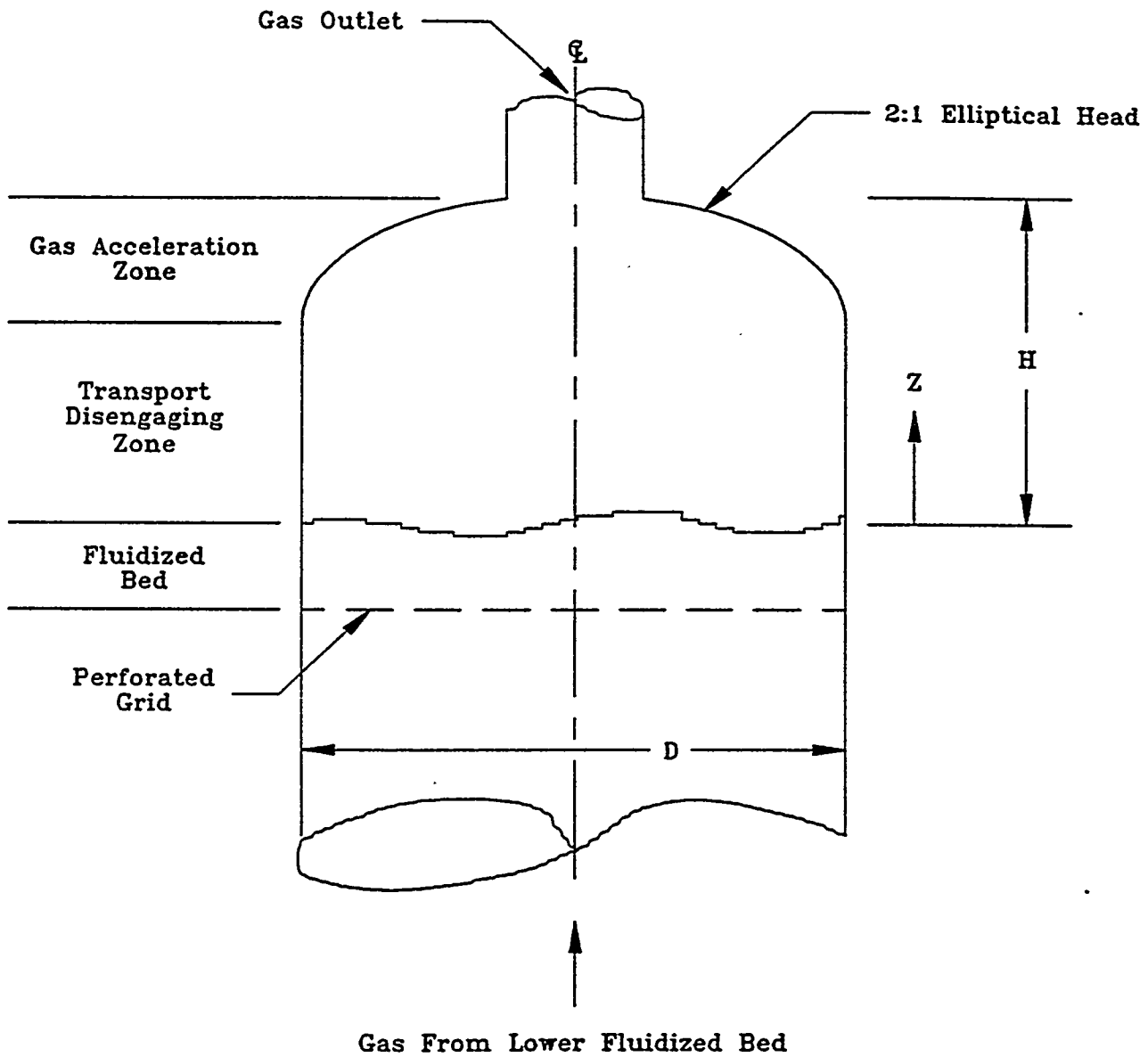
Flow modelling was performed using a commercial computational fluid dynamics (CFD) software program called PHOENICS. The effect of H and D on the axial velocity at the vessel centerline is indicated on Figure 3-8 and Figure 3-9. The centerline velocity was chosen for analysis since it is the maximum velocity at a specific Z position and consequently, the vessel centerline is the location where sorbent carryover would first occur. As seen from Figure 3-7 and Figure 3-8, as H increases for a fixed D, the velocity increase with increasing Z becomes more gradual. As expected, increasing H will decrease the velocity in the transport disengaging zone and minimize carryover.

**Figure 3-5. Comparison of the Simulated and  
Measured Sorbent Temperature at the  
Regenerator Sorbent Inlet**



**Figure 3-6. Comparison of the Simulated and Measured Sorbent Temperature at the Regenerator Sorbent Inlet**





**Figure 3-7. Upper Section of Fluid-Bed Vessel**

Figure 3-8. Centerline Velocity versus Height

Above Top Fluid Bed

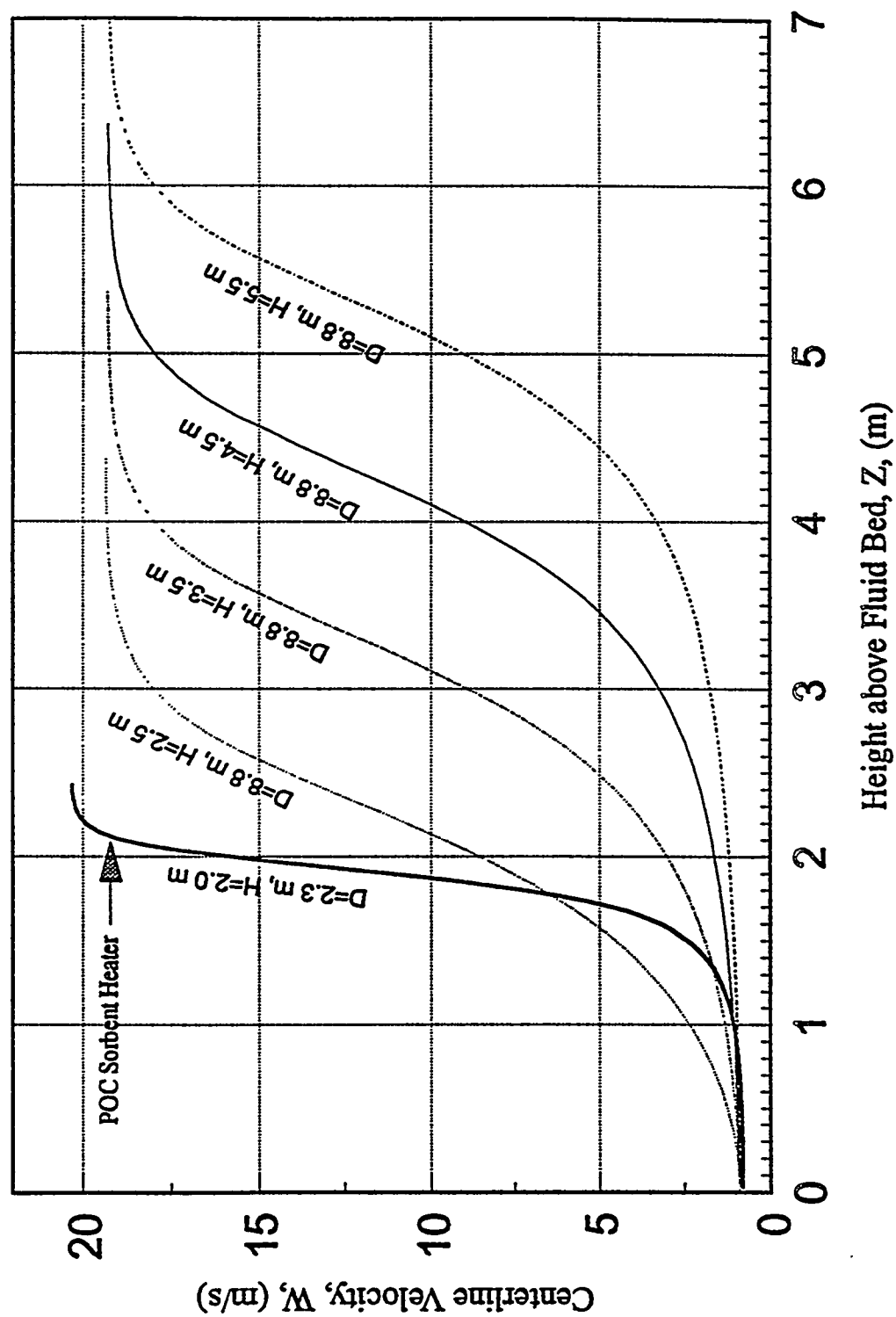
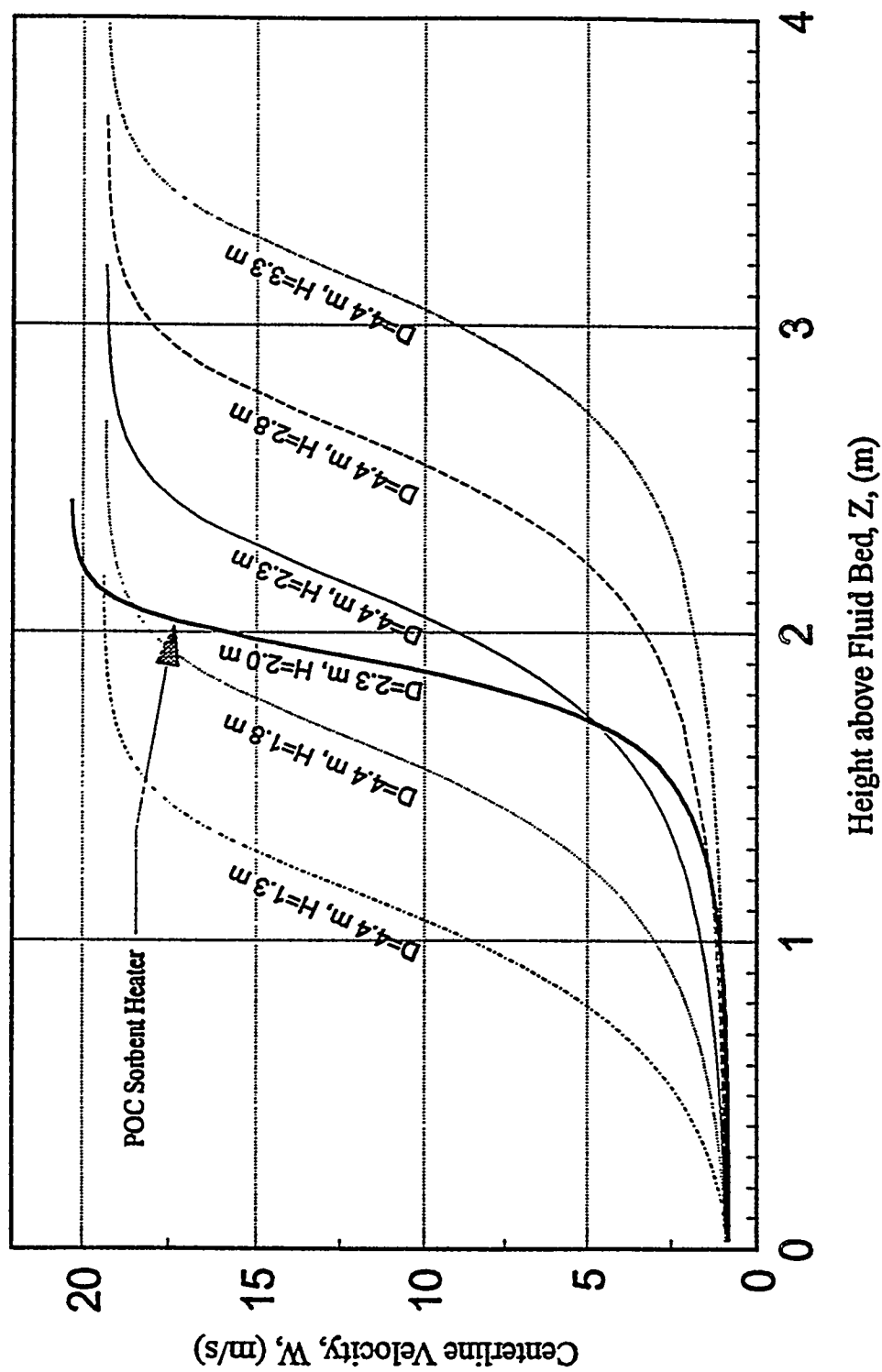


Figure 3-9. Centerline Velocity versus Height

Above Top Fluid Bed



A judgement is required to specify what magnitude of gas acceleration can be tolerated in the transport disengaging zone. To assist in making this judgement, the calculated centerline velocities for the POC sorbent heater are also shown on Figures 3-7 and 3-8. The POC sorbent heater is used as a reference since no sorbent carryover was observed during POC operations. From Figure 3-7 it is seen for a transport disengaging height of 1.5 m, H for an 8.8 m diameter vessel must be approximately 3.5 m to match the centerline velocity of the POC sorbent heater at the transport disengaging height. Similarly by using Figure 3-8, H for a 4.4 m diameter vessel must be approximately 2.5 m. Values of H greater than 3.5 and 2.5 m respectively would result in a more conservative design.

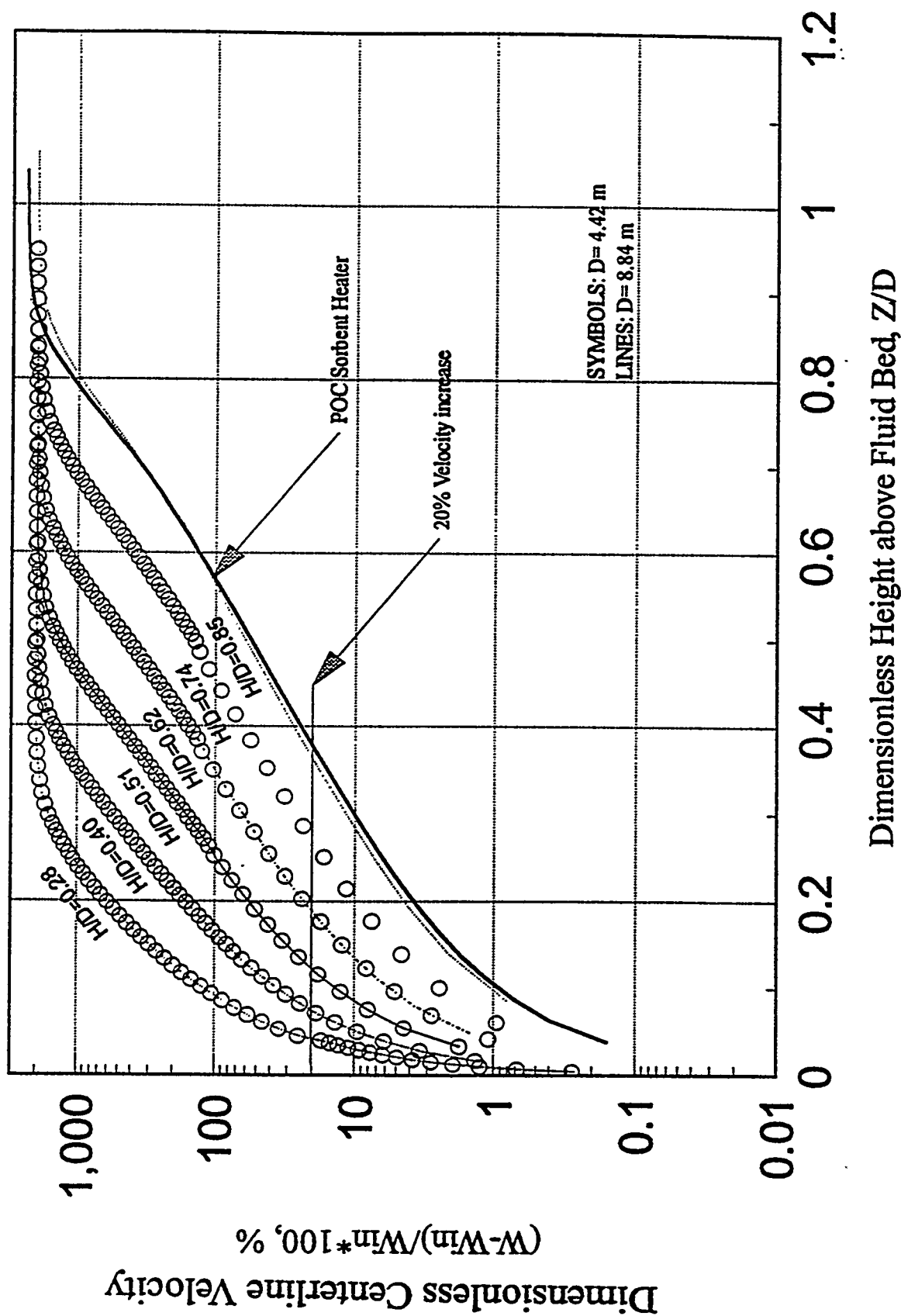
To develop a generic design procedure, the data from Figures 3-7 and 3-8 were plotted in dimensionless form on Figure 3-10 where  $W$  is the centerline velocity and  $W_{in}$  is the inlet velocity. It is seen that when presented in dimensionless form, the curves for a specific  $H/D$  are similar for different vessel diameters. A standard procedure for the vessel design is presented below.

1. Calculate the required vessel diameter ( $D$ ) for given gas velocity and temperature.
2. Determine the TDH from the lab test data for given bed depth and gas velocity.
3. Calculate the dimensionless  $[TDH] = TDH/D$ .
4. For the dimensionless  $[TDH]$  (normalized height above grid) and specified percentage velocity increase, e.g. 20%, determine  $H/D$  by interpolation from Figure 3-9. Based on POC data, a certain percentage velocity increase, e.g. 20%, within the TDH is acceptable.
5. Calculate  $H$  for given diameter  $D$ .

To demonstrate the above procedure using a specific example, the current design of the demonstration plant's sorbent heater will be used.

1. The top stage of the heater is 30.1b feet in diameter.
2. For the 1.33 ft settled bed depth, the required TDH is 7.1 ft.
3. The dimensionless TDH,  $[TDH]$ , is  $7.1/30.6=0.23$ .
4. From Figure 3-9, for  $Z/D=0.23$ , and a 20% velocity increase,  $H/D$  equals approximately 0.74.
5. The gas outlet nozzle should be located 22.6 ft above the top of the fluid bed, or 24 ft above the grid. Because the sorbent heater has a 2:1 elliptical head, the

**Figure 3-10. Dimensionless Centerline Velocity  
versus Dimensionless Height Above Fluid Bed**



head depth is 7.65 ft. Thus the cylindrical section of the vessel should extend approximately 16.35 ft above the top grid.

#### ***3.5.4 SO<sub>2</sub> Production Processes***

The NOXSO process has the flexibility to produce elemental sulfur, sulfuric acid, or liquid sulfur dioxide as a process by-product. Because of market sizes and the relative complexity of making the different products, elemental sulfur is generally the product of choice. However, for the demonstration project at AGC's Warrick Plant, the sulfur will be further processed to generate liquid sulfur dioxide to provide an adequate revenue stream to finance the project.

There are two primary methods to make sulfur dioxide from elemental sulfur, and there are several variations to each primary method. The two primary methods involve burning the sulfur in either air or oxygen. If sulfur is burned in air, the resulting stream contains about 16% SO<sub>2</sub> with the balance being mainly nitrogen, oxygen and water from the combustion air. The SO<sub>2</sub> must then be separated from the other combustion gases. This is typically done by stripping the SO<sub>2</sub> from the combustion gases with either water or an organic solvent like dimethylaniline or propylene carbonate. The SO<sub>2</sub> rich solvent is then regenerated for reuse, liberating the SO<sub>2</sub> in the process. If sulfur is burned in pure oxygen, the combustion gas contains pure SO<sub>2</sub> and no further processing is required. However, depending on the source of O<sub>2</sub>, the gas stream may contain some impurities and require further processing. The two primary methods for making liquid SO<sub>2</sub> are discussed in more detail below.

##### **Burn in Air with Water Stripping**

In the burn in air process, molten sulfur is pumped to the sulfur burner where it is atomized prior to combustion using compressed air. The exit gas from the sulfur burner typically contains 16-18% sulfur dioxide. A waste heat boiler is located downstream of the sulfur burner and the steam generated there is used elsewhere in the process. After the waste heat boiler, the process gas is further cooled in a series of two water spray cooling towers. The first tower is ceramic brick lined while the second tower is made of stainless steel. Water exiting the secondary cooling tower is recirculated to both the primary and secondary cooling towers. The portion of the water recycled to the secondary tower first passes through a heat exchanger where cooling water is used to reduce the process water temperature. Water recycled to the primary cooling tower bypasses the heat exchanger. The gas from the waste heat boiler flows cocurrent with the

water sprays in the primary cooling tower and countercurrent to the water sprays in the secondary cooling tower. The gas stream exiting the secondary cooling tower then passes to a compressor where the necessary pressure is developed to move the process gas through the rest of the system.

The remainder of the system involves stripping the  $\text{SO}_2$  from the process gas stream with water and then recovering the  $\text{SO}_2$  absorbed in the water. From the compressor, the process gas stream flows through two absorption towers in series. The gas stream exiting the second absorption tower still contains small amounts of  $\text{SO}_2$  and must be scrubbed before venting. A caustic scrubbing tower is used for this purpose and the exit gas from the scrubber typically contains less than 50 ppm of  $\text{SO}_2$ .

Fresh make-up water and water recovered from the stripping tower pass first through a heat exchanger where cooling water is used to reduce the temperature, and then to the second of the two absorption towers. Water exiting the second absorption tower is pumped to the first absorption tower. The  $\text{SO}_2$  rich liquor exiting the first absorption tower passes through two heat exchangers before going to the stripping tower for regeneration. The two heat exchangers use the hot stripped liquid exiting the stripping tower to heat the  $\text{SO}_2$  rich liquor prior to its entering the stripping tower. The hot regenerated liquid from the stripping tower after passing through the two heat exchangers combines with the fresh make-up water as discussed above.

In the stripping tower, steam is used to heat the  $\text{SO}_2$  rich liquor and liberate  $\text{SO}_2$  from the stream. The concentrated  $\text{SO}_2$  stream exiting the stripping tower still contains water and must be dried. This stream first passes through a condensing tower where cold water is used to reduce the  $\text{SO}_2$  containing stream temperature and condense most of the water in the stream. The remaining water is removed using three packed towers in series with a 93%  $\text{H}_2\text{SO}_4$  stream used as a desiccant. The spent acid (70%  $\text{H}_2\text{SO}_4$ ) must then be sent to an acid reclamation unit or a licensed disposal facility. The pure  $\text{SO}_2$  stream from the third drying tower passes through a compressor where the pressure is increased to about 80 psig and then passes through a condenser. Cooling water is used in the condenser, and liquid  $\text{SO}_2$  is accumulated in a collector tank before being pumped to storage.

## **Burn in Air with Sorbent Stripping**

The solvent based processes are nearly identical to the one described above with an organic solvent used instead of water as the stripping liquid. The advantage to the solvent based processes is the higher solubility of  $\text{SO}_2$  in the organics. However, some of the solvent is inevitably discharged to atmosphere during the process and the increasingly strict environmental regulations make permitting issues more of a concern.

## **Burn in Oxygen**

The production of liquid  $\text{SO}_2$  by burning sulfur in oxygen is a simpler process than the burn in air process described above. However, a source of oxygen is required which offsets some of the cost savings derived from the simpler liquid  $\text{SO}_2$  process.

Molten sulfur in the reaction chamber is first heated to above the sulfur auto-ignition temperature using electric heaters and then oxygen is introduced through a sparger ring four to six feet below the sulfur level. Oxygen at 80 psig is introduced at a rate such that sulfur is always the reactant in excess. In this manner, formation of  $\text{SO}_3$  is avoided. The heat generated from the sulfur combustion is consumed by vaporizing excess sulfur in the combustion chamber. The exit stream, containing  $\text{SO}_2$  and sulfur vapor, then passes through a sulfur condenser. Cooling water is converted to steam in the condenser. Enough steam is generated here for all the process needs of the burn in oxygen system and additional steam is available for export.

The condensed sulfur is gravity fed to a storage tank, and the effluent gas from the condenser also passes through the sulfur storage tank. In the storage tank, the temperature of the effluent gas is further reduced condensing out more sulfur. The remaining sulfur vapor is removed from the process gas stream by using sacrificial condensers which can build up solid sulfur and be cleaned by switching from water to steam to melt the accumulated sulfur. The process is operated at about 80 psig so that the  $\text{SO}_2$  can be condensed simply by using water from a cooling tower as a heat transfer medium. The condensed product is accumulated in an  $\text{SO}_2$  surge tank and from there it is intermittently pumped to storage.

The  $\text{SO}_2$  condenser and surge tank each have a vent line for the non-condensibles. However, this stream contains  $\text{SO}_2$  and is therefore sent to a caustic scrubber before being vented. Sodium sulfite is formed in the caustic scrubber and can either be sold as a by-product or properly

disposed. There are two levels of oxygen purity available for a burn in oxygen plant. The volume of the non-condensable stream is dependent on which purity oxygen is used. Cryogenic distillation is the more expensive option to generate oxygen, but the purity is very high (99.5%). In addition, a pure nitrogen stream is obtained as an additional product from the distillation tower. If there are sufficient users of nitrogen in the vicinity of the burn in oxygen SO<sub>2</sub> plant, sales of nitrogen help to offset the higher costs. The other method of producing oxygen is pressure swing adsorption (PSA). This is a less expensive method, but the O<sub>2</sub> purity is only 90-94%. In addition, no N<sub>2</sub> is produced as a by-product. The choice between the two O<sub>2</sub> production methods as well as between the two SO<sub>2</sub> production methods is ultimately one of economics.

### **3.6 Plant Characterization**

No plant characterization activities were conducted this quarter, however with the selection of a new host site, this activity will be revived.

### **3.7 Site Survey/Geotechnical Investigation**

Site survey/geotechnical investigation activities are on hold until a plant general arrangement for Warrick is developed.

### **3.8 Permitting**

Permitting activities are on hold until the necessary design information is developed to prepare permit applications.

## **4 PLANS FOR NEXT QUARTER**

Obtain a letter of intent from a consumer for the purchase of the liquid SO<sub>2</sub> to be produced by the project. Negotiations with potential buyers will be conducted. Additionally, technical evaluations of the SO<sub>2</sub> production process options and how each option is impacted by site conditions will be evaluated.

Collect the necessary data and conduct pollutant dispersion modelling for the EIV. The draft EIV will be updated and distributed for comment.

Update the preliminary process flow diagrams, heat and material balances, piping and instrumentation diagrams, and equipment specifications for the host site at Warrick. Develop a site plan and general arrangement for the NOXSO plant.

Complete the centrifugal separator tests to determine the removal efficiency on attrited sorbent and fly ash particulates.

Complete the three-stage, transparent, fluid-bed, adsorber model and confirm adequate solids transport through the system. Use the information from the transparent model to design and construct a metallic model which can be used to conduct adsorption studies using synthetic flue gas.

Conduct a study to determine the required grid pressure drop to assure uniform gas distribution across the fluid-bed vessels. This study will be done using a computational fluid dynamics software program.